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Contract No. CPA 22-69-78

FEASIBILITY STUDY OF NEW SULFUR OXIDE CONTROL PROCESSES FOR APPLICATION TO SMELTERS AND POWER PLANTS

Part IV: The Wellman-Lord SO_2 Recovery Process for Application to Power Plant Flue Gases

Prepared for:

U.S. DEPARTMENT OF HEALTH, EDUCATION, AND WELFARE
NATIONAL AIR POLLUTION CONTROL ADMINISTRATION
DURHAM, NORTH CAROLINA

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By: KONRAD T. SEMRAU

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SRI Project PMU-7923

Approved:

N. K. HIESTER, *Director*
Physical Sciences (Materials)

C. J. COOK, *Executive Director*
Physical Sciences Division

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A block flow diagram was prepared.
Mass and volume flow balances were estimated. Engineering estimates of the capital investment and the total annual cost (including both fixed and variable charges) were prepared. These engineering estimates were applied to previously created model gas streams and the total annual cost for each was determined. From these annual costs, incremental production costs were estimated. The technical constraints on the application and operation were qualitatively appraised; the economic constraints were appraised (quantitatively, to the extent permitted by the data). The current state of development was assessed, identifying any technological deficiencies whose elimination might enhance applicability.

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FOREWORD

The final report for this study is presented in four separate and independent parts,

Part I : The Monsanto Cat-Ox Process for Application to Smelter Gases

Part II : The Wellman-Lord SO_2 Recovery Process for Application to Smelter Gases

Part III: The Monsanto Cat-Ox Process for Application to Power Plant Flue Gases

Part IV : The Wellman-Lord SO_2 Recovery Process for Application to Power Plant Flue Gases

Information for use in this study was supplied to Stanford Research Institute by Monsanto Company and Wellman-Lord, Inc. under terms of confidentiality agreements between the U. S. Department of Health, Education and Welfare, Stanford Research Institute, and each of the cooperating companies. In accordance with the agreements, Monsanto Company and Wellman-Lord, Inc. have reviewed and released the parts of the report dealing with their respective processes. The rights of prior review and release are designed solely to permit the cooperating companies to assure themselves that no proprietary or confidential data are being revealed; they are not intended to restrict Stanford Research Institute's rights and responsibilities to report its conclusions so long as there is no incidental disclosure of confidential information. Accordingly, the release of the reports by Monsanto and Wellman-Lord does not imply that these companies necessarily concur in all or any of the opinions, judgments, or interpretations of fact expressed by the author, who assumes sole responsibility for the report content.

This report was furnished to the
Air Pollution Control Office by
the Stanford Research Institute
in fulfillment of Contract No.
CPA 22-69-78.

I INTRODUCTION

Under the Systems Study for Control of Emissions -- Primary Nonferrous Smelting Industry (Contract No. PH 86-68-85), Arthur G. McKee & Company and its subcontractor, Stanford Research Institute, carried out evaluations of a number of sulfur oxide control processes as they might be applied to offgases from nonferrous smelting. To permit evaluation of the technical and economic feasibility of these control processes, a number of models of smelters were created. Stanford Research Institute carried out the studies necessary to determine the availability of markets for sulfur by-products open to smelters in various areas, and the allowable production costs that the smelters would have to attain in order to break even on the sulfur recovery operations.

The Division of Process Control Engineering of the National Air Pollution Control Administration (DPCE-NAPCA) desires to extend the usefulness of the foregoing study by adding to it technical and economic evaluations of new and potentially promising sulfur oxide control processes. It also wishes to evaluate the same new processes for application to power plants. Completion of these preliminary evaluations of the processes will help determine their potential commercial acceptability.

DPCE-NAPCA has a specific interest in at least two control processes being offered commercially, the Monsanto Cat-Ox process and the Wellman-Lord SO_2 Recovery process. However, both processes are proprietary, and, as a matter of policy, DPCE-NAPCA does not wish to obtain proprietary and confidential information on the processes. It does, nevertheless, wish to obtain evaluations in nonconfidential terms. Broadly, DPCE-NAPCA wishes to obtain estimates of the capital and annual costs of the control systems for each of the assumed applications, together with appraisals of the technical constraints on each process and of the current states of development of the processes.

Stanford Research Institute was requested by DPCE-NAPCA to carry out evaluations of the processes under the terms of confidentiality agreements between the Department of Health, Education, and Welfare, the owners of the proprietary processes, and Stanford Research Institute. SRI, acting as a disinterested third party, was to make analyses of the processes using information obtained from Monsanto Company and Wellman-Lord, Inc., and to report the results to DPCE-NAPCA without compromising any of the Monsanto or Wellman-Lord confidential data.

Because of the requirements of confidentiality, it is not permissible to describe certain features of the Cat-Ox and Wellman-Lord processes. The corresponding portions of the systems have had to be represented only in terms of their general functions, and SRI's evaluation of these portions has had to be presented in the form of conclusions without supporting data or reasoning. In other instances, the parts of the systems could be described in general, but specific details and design parameters could not be revealed.

Within the scope of the present project it would obviously have been impossible to inspect and evaluate independently all the company records and design data even had the cooperating companies been requested to permit this and had they acceded to the request. The author of this report, who also conducted the study, evaluated the information provided at his request, using his own knowledge and relevant data from the literature and other available sources. Whenever apparent discrepancies or uncertainties were noted in the information, efforts were made to secure verification or clarification from the companies. In instances where resolution of questions was not possible, or the information required proved to be simply unavailable, the author employed his best judgment.

Throughout the following sections of this report, information for which other sources are not specifically cited was generally obtained from the cooperating companies and accepted by the author either because it could be verified from other sources or because it appeared reasonable. In other instances, information or estimates were provided by the companies that could not be verified independently or judged for reasonableness; in such cases, the companies have been specifically cited as the

sources. In still other instances, the author did not accept the information or estimates provided and in some cases substituted his own; such cases have also been specifically noted.

The cooperating companies, at their own option and through substantial efforts, provided the basic capital cost estimates for the model control systems and the information for estimation of operating and maintenance costs. The author in this case acted as a reviewer rather than as an estimator. The estimates were checked for reasonableness and for possible errors or omissions. For some components and cost factors, the author modified the estimates, or substituted others of his own where he judged them to be more appropriate than those supplied to him. The author also prepared cost estimates for some auxiliary systems, using separate data sources.

By specification of the power plant and smelter models, and by review of the results, an effort was made to ensure that the cost estimates for both control systems were made on strictly comparable bases. Although it is unlikely that this objective has been met fully, the deviations are probably within the precision of the estimates themselves.

For convenience, and at the request of DPCE-NAPCA, this final report is presented in four separate and independent parts. This part deals only with the Wellman-Lord SO_2 Recovery process as applied to control of sulfur oxides in the flue gases from power plants.

II OBJECTIVES

The objectives of the part of the study covered by this report are as follows:

1. To prepare block flow diagrams of the Wellman-Lord system showing its configuration and its relation to the boiler plant in which the sulfur oxides are generated.

2. To present estimated mass and volume flow balances for the Wellman-Lord system.

3. To prepare preliminary engineering estimates of the capital investment and the total annual cost (including both fixed and variable charges) for the Wellman-Lord system. From the estimate of total annual cost, secondary estimates are to be made of the corresponding incremental cost of producing electricity, both on the gross basis (without allowance for by-product recovery credits) and on the net basis (with allowance for by-product recovery credits).

4. To make a qualitative appraisal of technical constraints on the application and operation of the control system.

5. To appraise (quantitatively, to the extent permitted by available data) the economic constraints on the application of the control system.

6. To assess the current state of development of the Wellman-Lord system, identifying any technological deficiencies whose elimination might enhance the applicability of the system to power plants.

The accomplishment of the objectives is subject to any restrictions that may be imposed under the terms of the confidentiality agreement between the Government, Wellman-Lord, Inc. and Stanford Research Institute.

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III SUMMARY

The Wellman-Lord SO_2 Recovery process is a cyclic absorption-desorption process for producing concentrated sulfur dioxide from waste gas streams, such as power plant flue gas and smelter offgases. It is adaptable to use on gas streams containing sulfur dioxide at both low and relatively high concentrations.

The basic Wellman-Lord system is made up of two basically different parts: (1) the absorption section, in which the sulfur dioxide is removed from the flue gas, and (2) the chemical recovery section, in which the concentrated sulfur dioxide is recovered from the spent absorbent and the absorbent is regenerated for return to the absorption section. The absorbent is a solution of sodium sulfite. In the absorption cycle, the sulfite ion reacts with sulfur dioxide, forming bisulfite. In the recovery cycle, the reaction is reversed by application of heat, releasing sulfur dioxide and regenerating the sulfite.

The absorption step is conventional and has been employed commercially as well as studied experimentally. The regeneration step apparently represents an innovation; it is intended to reduce the amount of steam required where regeneration is accomplished by direct stripping of the bisulfite solution with steam. The Wellman-Lord regeneration (or chemical recovery) system appears to be capable of operating with a steam demand in the range of about one-half to one-third that required in a conventional steam stripping system. The regeneration equipment is relatively expensive, but not sufficiently so for the associated fixed charges to offset the advantage of the reduction in steam consumption. Since the regeneration system has not yet received patent protection, it cannot be described at this time.

It is essential to the operation of the Wellman-Lord system that the flue gas entering the absorber be as nearly free of fly ash as practical. Fly ash collected in the absorbent solution may produce a

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variety of mechanical and chemical problems. The electrostatic fly ash precipitator removes the bulk of the fly ash, but a prescrubber is also installed immediately upstream from the absorber. The prescrubber collects most of the residual fly ash that escapes collection by the electrostatic precipitator, and also cools and humidifies the flue gas.

The Wellman-Lord system treated in this study is essentially a conceptual design. Components of the system have been tested individually, but a complete, integrated system has not yet been operated. Another version of the Wellman-Lord process, using potassium sulfite solution as the absorbent, was tested in a pilot plant and in a demonstration plant but was generally unsuccessful. The regeneration system was basically different from that used with the sodium-base absorbent, and it did not attain the steam economy originally anticipated. Many of the problems encountered resulted from entry of fly ash into the system; the prescrubber used was inadequate in efficiency and plugged rapidly with fly ash.

The Wellman-Lord system as designed for use with the sodium-base absorbent appears to be technically feasible and to avoid most of the problems that were encountered in the demonstration plant using the potassium-base absorbent system.

In the Wellman-Lord process, as in all similar processes, some portion of the sulfite (or sulfur dioxide) is oxidized to sulfate from which the sulfur dioxide cannot be regenerated. The sulfate must be purged from the system. The fraction of the sulfur dioxide that is oxidized has not been established. Wellman-Lord originally estimated that the amount would be very low and provided no system for recovering the sodium base from the purge stream of absorbent solution, which was expected to be small. The author anticipated that the amount of sulfur dioxide oxidized will be much larger than that originally projected by Wellman-Lord, and assumed the amount to be 1.0 percent of that absorbed, although the actual percentage is unpredictable. He therefore assumed that the purged absorbent solution would be recausticized with lime to recover the sodium base.

The purging of the sulfate results in a loss of sulfur dioxide in the sodium sulfite present in the purge stream. If as much as 4 to 6 percent of the sulfur dioxide should be oxidized, a substantial fraction of the total sulfur dioxide will be lost in the purge, and a system for separating the sodium sulfate and sulfite will be needed.

The product of the Wellman-Lord system, concentrated sulfur dioxide, has only very limited markets. Hence, it will in general be necessary either to convert the sulfur dioxide to sulfuric acid in an auxiliary contact acid plant, or to reduce it to elemental sulfur in an auxiliary reduction plant.

The absorption tower of the Wellman-Lord system is designed to remove 90 percent of the sulfur dioxide from the flue gas. However, some of the recovered sulfur dioxide is subsequently lost to the atmosphere from the sulfuric acid plant or the reduction plant used to convert the sulfur dioxide. The overall sulfur dioxide emission control efficiency is estimated to be 88 percent for the combination of the Wellman-Lord system and acid plant, and 84 percent for the Wellman-Lord system and reduction plant. The sulfur dioxide emissions from the assumed auxiliary conversion plants were estimated by the author. Various potential methods might be used to recover part of these emissions. For example, Wellman-Lord absorbers might be applied to the tail gases, with the spent absorbent being returned to the central chemical recovery section that serves the main absorption section. The acid plant might employ the double-contact process to increase the sulfur dioxide conversion.

To permit estimates of capital and annual costs of the Cat-Ox and Wellman-Lord SO_2 Recovery systems, two models of hypothetical power plants were created. (Details are presented in Appendix A.) The first (Model A) is an existing 500-megawatt plant located in central Pennsylvania (Altoona). The second (Model B) is a new 1000-megawatt plant located on a navigable river in the Midwest (Cairo, Illinois). Estimates of annual cost were based on a plant life of 20 years and on an annual operating time of 7000 hours (load factor 80 percent). The estimated prices for by-product sulfuric acid are based on an assumed Gulf Coast price for sulfur of $30/long ton, which is probably as high as can be anticipated in the period up to 1975.

Cost estimates were made for Wellman-Lord SO_2 Recovery systems to be applied to each hypothetical power plant, and to each recovery plant were added alternative auxiliary contact sulfuric acid plants and sulfur dioxide reduction plants. The costs for the auxiliary plants were estimated by the author. The estimates of capital and operating costs for the four control system models are summarized in Tables I through IV. The annual costs and the equivalent incremental costs of electricity generated and fuel used are presented both before and after allowance for sulfur by-product credits. In the case of the Model A control system producing acid, two acid prices are presented in estimating the by-product credit; the lower of the two is applicable in the event that it should be necessary to displace existing captive acid producers in the Pittsburgh area, where the acid from the Altoona plant would have to be sold.

Although the cost of producing elemental sulfur exceeds that of producing sulfuric acid, disposing of the acid would often be difficult in actual cases, whereas the sulfur can be shipped greater distances or be readily stored.

Table I

POWER PLANT MODEL A
SUMMARY OF ESTIMATED COSTS FOR WELLMAN-LORD SO_2 RECOVERY SYSTEM
WITH CONTACT SULFURIC ACID PLANT

Item	Cost Basis	Cost or Credit
Capital Investment		
A. Wellman-Lord System	$	7,240,000
B. Contact Acid Plant	$	790,000
Total	$	8,030,000
	$/kw	16.06
Gross Costs Before Credits	$/yr	2,974,165
	Mills/kwh	0.850
	¢/million Btu	8.50
	$/ton of coal	2.26
Credits for Sulfuric Acid 95,025 tons/yr (100% basis)	$12.00/ton	1,140,300
	$ 7.00/ton	665,175
Net Costs After Credits		
A. Acid Price = $12/ton	$/yr	1,833,865
	Mills/kwh	0.524
	¢/million Btu	5.24
	$/ton of coal	1.40
B. Acid Price = $7/ton	$/yr	2,308,990
	Mills/kwh	0.660
	¢/million Btu	6.60
	$/ton of coal	1.76

Table II

POWER PLANT MODEL A
SUMMARY OF ESTIMATED COSTS FOR WELLMAN-LORD SO_2 RECOVERY SYSTEM
WITH SULFUR DIOXIDE REDUCTION PLANT

Item	Cost Basis	Cost or Credit
Capital Investment		
A. Wellman-Lord System	$	7,240,000
B. SO_2 Reduction Plant	$	1,040,000
Total	$	8,280,000
	$/kw	16.56
Gross Costs Before Credits	$/yr	3,174,135
	Mills/kwh	0.907
	¢/million Btu	9.07
	$/ton of coal	2.42
Credits for Elemental Sulfur 29,330 short tons/yr	$33.00/long ton	864,190
Net Costs After Credits	$/yr	2,309,945
	Mills/kwh	0.660
	¢/million Btu	6.60
	$/ton of coal	1.76

Table III

POWER PLANT MODEL B
SUMMARY OF ESTIMATED COSTS FOR WELLMAN-LORD SO_2 RECOVERY SYSTEM
WITH CONTACT SULFURIC ACID PLANT

Item	Cost Basis	Cost or Credit
Capital Investment		
A. Wellman-Lord System	$	12,010,000
B. Contact Acid Plant	$	1,200,000
Total	$	13,210,000
	$/kw	13.21
Gross Costs Before Credits	$/yr	4,589,750
	Mills/kwh	0.656
	¢/million Btu	6.83
	$/ton of coal	1.82
Credits for Sulfuric Acid 182,350 tons/yr (100% basis)	$10.00/ton	1,823,500
Net Costs After Credits	$/yr	2,766,250
	Mills/kwh	0.395
	¢/million Btu	4.12
	$/ton of coal	1.10

Table IV

POWER PLANT MODEL B
SUMMARY OF ESTIMATED COSTS FOR WELLMAN-LORD SO_2 RECOVERY SYSTEM
WITH SULFUR DIOXIDE REDUCTION PLANT

Item	Cost Basis	Cost or Credit
Capital Investment		
A. Wellman-Lord System	$	12,010,000
B. SO_2 Reduction Plant	$	1,520,000
Total	$	13,530,000
	$/kw	13.53
Gross Costs Before Credits	$/yr	4,886,020
	Mills/kwh	0.698
	¢/million Btu	7.27
	$/ton of coal	1.94
Credits for Elemental Sulfur 56,280 short tons/yr	$33.00/long ton	1,658,250
Net Costs After Credits	$/yr	3,227,770
	Mills/kwh	0.461
	¢/million Btu	4.80
	$/ton of coal	1.28

IV PROCEDURES

A. Formulation of Models

The formulation of the models of the hypothetical power plants was based first on conditions specified by DPCE-NAPCA. Part of the remaining model conditions were derived from the NAPCA specifications so as to be consistent with the latter. Other conditions had to be selected on a relatively arbitrary basis, and where this was necessary, guidance was obtained from the literature or from discussions with informed individuals. In other cases, conditions had to be selected on completely arbitrary bases, and the author used his own experience and best judgment. The complete models are presented in Appendix A.

The basic premise employed throughout was that all estimates of the costs of sulfur oxides control should be made in relation to a base level represented by the equivalent conventional plant without sulfur oxides emission controls. Emission control was to be charged with whatever costs were incurred above those of the conventional, or base-level, plant. Similarly, it was to be credited with any savings from the cost of the base-level plant, or with any income derived from sale of by-products.

If the sulfur oxide control system included components common to the base-level plant, it was to be charged only with the increment in the costs of these components over those of the corresponding items in the base-level plant (or credited with the difference if the items in the conventional plant were more expensive).

The sulfur oxide control system was to be charged with the cost of power required to move the flue gas through any parts of the system specific to emission control. It was also to be charged with a proportionate share of the capital cost of the fan and motor if the latter also supplied draft for the rest of the boiler system.

The model formulations were circulated to DPCE-NAPCA and to Monsanto Company and Wellman-Lord, Inc. for review before adoption.

B. Cost Factors

The factors used in making cost estimates, and the bases on which they were adopted, are presented in Appendix B. Salaries, payroll benefits, and overhead were taken to be the same as those used in the previous study of the nonferrous smelting industry.[15]

Estimates of the prices that might be obtained for sulfur by-products were made by Stanford Research Institute, and are presented in Appendix D. The choices of the specific sites for the hypothetical plants (within the general areas specified by DPCE-NAPCA) were made to facilitate estimation of definite prices.

C. Preparation of Technical Data and Cost Estimates

Wellman-Lord, Inc. prepared the technical designs for the model control systems, based on the conditions formulated by SRI, and estimated the capital investments and the utilities and maintenance requirements. The author reviewed these estimates and accepted most of them. The cost breakdowns for individual components of the system were supplied by Wellman-Lord for review. Most of the components consisted of conventional items of chemical process equipment, and the costs were estimated by Wellman-Lord on the basis of quotations from vendors. The author surveyed these estimates for general reasonableness and consistency with the model specifications, but did not attempt further verification.

Some of the original estimates of system requirements or costs made by Wellman-Lord were not accepted, and the author supplied his own estimates. Such instances are noted in the following sections of this report. The changes were discussed with Wellman-Lord, and differences were generally resolved.

The cost estimates for the auxiliary conversion plants (contact sulfuric acid and sulfur dioxide reduction) were made by the author from published data.[6,28] (See also Appendixes D and E of Part II of this report.)

V PROCESS DESCRIPTION

The Wellman-Lord SO_2 Recovery process is a cyclic absorption-desorption process for producing concentrated sulfur dioxide from waste gas streams, such as power plant flue gas and smelter offgases. It is adaptable to use on gas streams containing sulfur dioxide at both low and relatively high concentrations. Its product, concentrated sulfur dioxide, has very limited markets, but can be used in sulfuric acid manufacture or be reduced to elemental sulfur. Therefore, the Wellman-Lord process potentially has two principal uses: (1) concentration of sulfur dioxide from dilute gas streams (those containing less than about 2 to 3 percent of sulfur dioxide) for use in sulfuric acid plants, and (2) concentration of sulfur dioxide from both dilute and rich gas streams for subsequent reduction to elemental sulfur. If sulfuric acid is the desired product, and the gas stream contains at least 3.5 to 4 percent of sulfur dioxide, it is practical to treat the gas directly in a contact sulfuric acid plant.[8]

The absorption step in the Wellman-Lord process cycle is conventional, and in closely related forms has been used commercially and been studied widely on an experimental basis by numerous investigators. Wellman-Lord, Inc. believes that some elements of the absorber design are novel and is seeking patent protection for them. Although the writer questions -- on purely technical grounds -- whether the design features are actually novel, it is not critical to the present evaluation of the basic process whether the claimed innovations in equipment are real or not.

The absorbent consists of a solution of sodium or potassium sulfite. In the absorption cycle, the sulfite ion reacts with sulfur dioxide as indicated forming bisulfite:

$$SO_3^{=} + SO_2 + H_2O = 2\ HSO_3^{-}$$

In the desorption cycle, the reaction is reversed by application of heat releasing sulfur dioxide and regenerating the sulfite:

$$2\ HSO_3^- = SO_3^= + H_2O + SO_2$$

The absorption step has been employed commercially in the pulp and paper industry (usually with sodium and ammonia as bases) for production of cooking liquor;[1,5,18] in this case no regeneration step is employed. The complete cycle, including the desorption step, has been studied experimentally[12,13] and employed occasionally on a limited commercial scale.[12] However, the major economic limitation has been the quantity of steam required for the absorbent regeneration. In the conventional approach to regeneration,[13] the rich absorbent (containing a large fraction of bisulfite) is stripped of sulfur dioxide with steam in a countercurrent tower. At the bottom of the tower where the lean stripped absorbent solution emerges, the bisulfite has been largely converted to sulfite and the back pressure of sulfur dioxide above the solution is low. The practical limits of regeneration of the absorbent are set by the required tower height and the quantity of steam.

Various attempts have been made to avoid the steam stripping of the absorbent. Johnstone,[14] for example, proposed introducing a chemical regeneration cycle. The Wellman-Lord process employs crystallization of salts from the absorbent as part of a procedure for reducing the steam requirements for regeneration. Various proposed forms of the process are described in Wellman-Lord patents.[3,16,17] One, which uses the potassium-base salts, is described broadly in a Belgian patent,[3] and more specifically in a paper by Watt.[27] It was studied on a pilot plant scale[26] and was later investigated in a demonstration plant.[27] The anticipated steam economy was not attained, and development has been shifted to another form of the process, which employs the sodium-base salts.

The sodium-base process, which is used as the basis for the present study, has a regeneration process that is essentially different from that employed where the potassium-base salts are used. It is the regeneration process that constitutes the apparently novel feature of the

Wellman-Lord system. Applications have been made for patents, but patent protection has not yet been received. Consequently, it is not permissible at this time to describe the basic regeneration process or the system for carrying it out. The corresponding portions of the regeneration system are therefore omitted from the flowsheet, Fig. 1.

The system considered in the present study has not been operated in a complete, integrated system, but experiments have been carried out with individual components. It is essentially a conceptual design backed by relevant experience with the potassium-base pilot plant and demonstration plant and by data from the tests of system components. A commercial plant using the process is to go into service in 1970 on the tail gas from a contact sulfuric acid plant.

The regeneration system devised by Wellman-Lord avoids the direct steam stripping of the absorbent in a stripping column, and reduces the steam requirement substantially below that of the stripper. During regeneration, the amount of sulfur dioxide recovered per unit quantity of absorbent circulated depends upon the concentration of the solution and upon the fraction of the salt that is initially bisulfite. Hence, attaining maximum steam economy (the least consumption of steam per unit quantity of sulfur dioxide recovered) requires that the rich absorbent leaving the absorption process be as concentrated as possible, and that the conversion of the sulfite to bisulfite also be as nearly complete as possible.

The upper limit of concentration of the sodium sulfite in the lean absorbent solution fed to the absorber is set by the solubility of the salt. As the sulfite is converted to the more soluble bisulfite by reaction with sulfur dioxide (the actual crystalline compound is the pyrosulfite instead of the bisulfite), the solution becomes less nearly saturated. Hence, a portion of the water in the solution can be evaporated into the gas stream passing through the tower. However, a practical limit on the amount of water removed in this manner is imposed by the need to avoid bringing the rich absorbent solution to saturation with respect to sodium pyrosulfite. If saturation is reached, plugging of parts of the absorber and absorbent outlet piping by pyrosulfite crystals may occur.

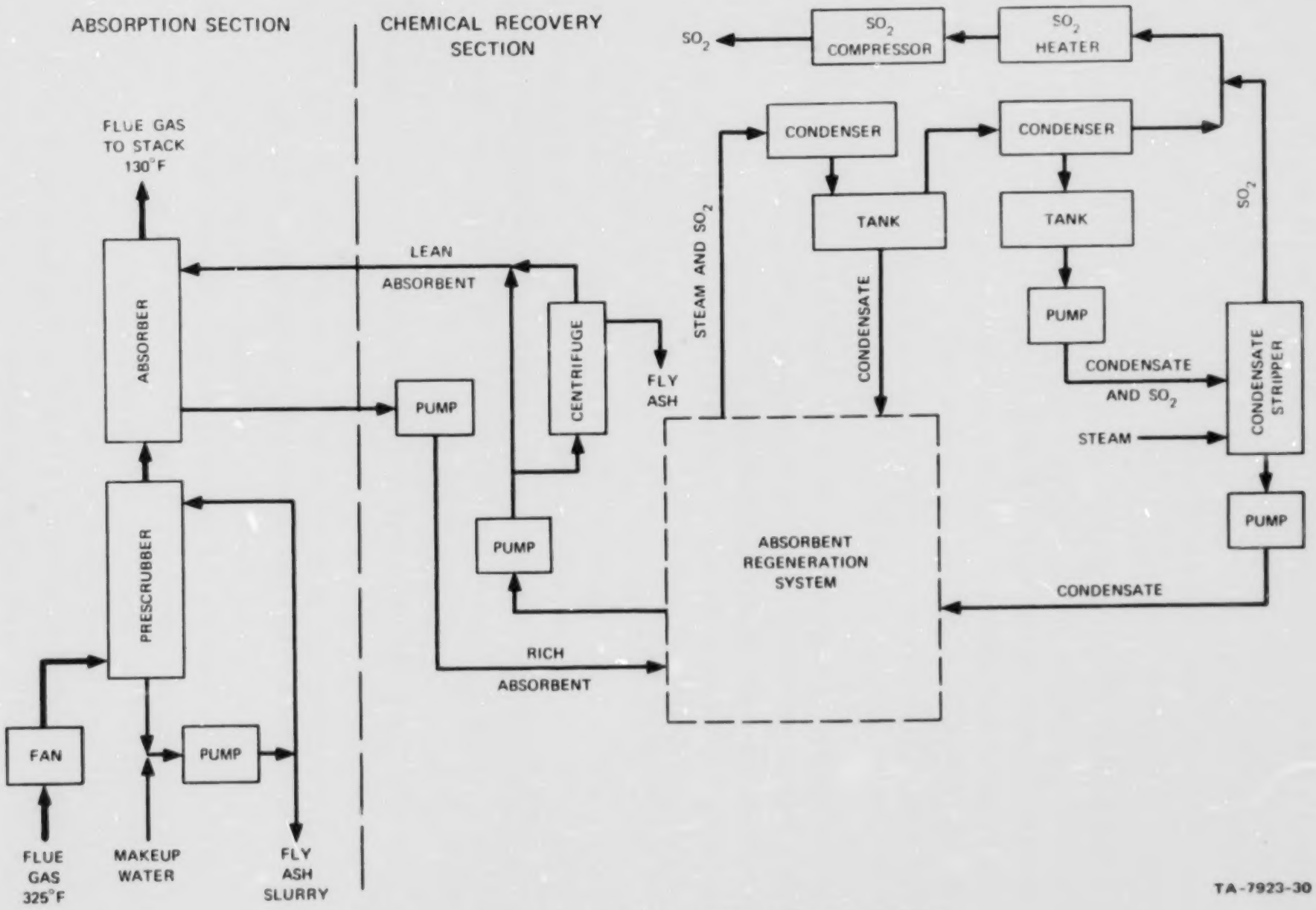

FIGURE 1 WELLMAN-LORD SO_2 RECOVERY SYSTEM FOR POWER PLANT FLUE GAS

The circulation rate of the absorbent is set by the quantity of sulfur dioxide to be removed from the gas stream. If the gas being treated is relatively rich in sulfur dioxide (as, for instance, a smelter gas), it is possible in a countercurrent absorption tower to convert somewhat more of the sulfite to bisulfite than if the gas is more dilute; a higher back pressure of sulfur dioxide above the solution leaving the tower is permissible because the partial pressure of sulfur dioxide in the entering gas stream is also higher. However, the additional fractional conversion of the sulfite is not large,[13] and the gain in capacity of the absorbent is ignored in conservative design of a system. The sulfur dioxide capacity is therefore taken as constant regardless of the composition of the gas stream treated.

The steam economy of the system can be improved by staging the regeneration process, but the economic limits of this approach are set by the concomitant increase in capital investment. The optimum balance between operating costs and depreciation must therefore be determined.

In the Wellman-Lord process, as in all similar processes employing the sulfite-bisulfite absorption and recovery cycle (or even absorption without regeneration), some portion of the sulfite is oxidized to sulfate, from which the sulfur dioxide cannot be regenerated. The sulfate must therefore be purged from the system. The fraction of the sulfite that may be oxidized in the Wellman-Lord system has not yet been established; it may, in fact, vary widely depending upon such factors as the oxygen content of the gas stream, the temperature, and the presence of contaminants that may act as oxidation catalysts. Wellman-Lord adds hydroquinone to the absorbent solution as an oxidation inhibitor, but the effectiveness of this measure is questionable. This point is discussed below.

The basic Wellman-Lord system is made up of two basically different parts, the absorption section and the chemical recovery section (see Fig. 1). The absorption section is composed only of the absorber, the prescrubber, and the fan. Its size, and hence its cost, is dependent almost solely upon the volume of gas handled, with only a minor cost

component related to the concentration of the sulfur dioxide in the gas. The size and cost of the chemical recovery section are dependent upon the total quantity of sulfur dioxide recovered. Since the quantity of sulfur dioxide absorbed per unit quantity of absorbent is constant, the total volume of absorbent circulated is directly proportional to the total quantity of sulfur dioxide.

Removal of particulate matter from the gas before the latter enters the absorber is a critical aspect of the operation of the Wellman-Lord process. (It will evidently be equally critical to any other cyclic process for concentration of sulfur dioxide.) Particulate matter entering the absorption system may produce mechanical problems in the chemical recovery section (e.g., plugging lines and wearing pumps) and may interfere in other ways with the operation of the system. It must be purged from the system by filtration or centrifuging and may carry with it occluded absorbent. If the filtered particulate is washed to recover the absorbent, the resulting dilution of the solution increases the steam requirements for the regeneration process. In addition, the particulate matter may include elements that will act as catalysts for the oxidation of the sulfite.

Acidic gases and vapors other than sulfur dioxide are also objectionable, since they will react with and consume absorbent. Some of the sulfur trioxide present in the flue gas will be collected in the absorber and form sodium sulfate, which must then be purged from the system along with the sulfate formed by the oxidation of part of the sulfite. Hydrogen chloride and hydrogen fluoride, which may also be present in small quantities, will form sodium chloride and sodium fluoride that must also eventually be purged from the absorbent. A small quantity of sodium nitrate may also be formed by the absorption of nitrogen oxides.

In the present study, the model power plants are assumed to have the same electrostatic fly ash precipitators as would the corresponding plants without sulfur oxide emission controls. In addition, the proposed Wellman-Lord system includes the prescrubber specified by Wellman-Lord. The prescrubber, which is only one of several potential alternative

types, is provided with trays for countercurrent gas-liquid contacting. It is actually constructed within the same shell as the absorber, although this arrangement is not indicated in Fig. 1 and is not essential, but only a matter of convenience. The gas pressure drop is 4 inches of water. Wellman-Lord anticipated that the prescrubber will remove up to about 90 percent of the fly ash present in the flue gas as it comes from the electrostatic precipitator. On the basis of scrubber performance on fly ash reported in the literature,[19,20] the anticipated efficiency appears reasonable. The water used for scrubbing is recirculated with makeup water provided to compensate for evaporation and the water in the fly ash slurry discharged to waste. Since the scrubbing water is recirculated, the scrubbed gas should approach saturation at its adiabatic saturation temperature, which is about 128°F. The cooling and humidification of the flue gas will also control the evaporation of water from the absorbent solution in the absorber.

In addition to collecting most of the fly ash that escapes the electrostatic precipitator, the prescrubber will also collect a portion of the sulfur trioxide. However, the sulfur trioxide will react with water vapor in the prescrubber to form a sulfuric acid aerosol, and the collection efficiency is not readily predictable; in this study the author assumed an efficiency of 50 percent. The prescrubber may remove a portion of any hydrogen chloride and hydrogen fluoride present in the flue gas, but the acidic, recirculated scrubber water will not be a favorable absorbent for these gases.

The absorber will also remove additional amounts of the fly ash and sulfuric acid mist. The author assumed an overall efficiency on fly ash (for the prescrubber and absorber) of 99.5 percent, which appears reasonable in view of the total pressure drop of 16 inches of water resulting from gas-liquid contacting. He also assumed an overall efficiency of 90 percent on the sulfuric acid mist, but this is much more speculative.

The flue gas leaving the prescrubber is forced up through the countercurrent absorbing tower where it contacts the absorbent solution on

a series of trays and the sulfur dioxide is absorbed. The treated gas leaving the top tray passes through an entrainment separator where entrained droplets of the absorbent solution are removed, and it is then discharged to the stack. Because the concentration of the sulfur dioxide is low, the liquid throughput is also relatively low and special provisions (not indicated in Fig. 1) are made to ensure good gas-liquid contact. The gas pressure drop across the absorber is about 16 inches of water, including the entrance and exit losses of both the prescrubber and absorber. The largest individual absorption towers are rated at 505,000 ACFM of gas, and hence both of the model power plants require multiple parallel towers (three in the Model A plant and six in the Model B plant).

In the Model B plant the fans and large pumps are driven by back-pressure steam turbines, whose exhaust supplies the low pressure steam needed in the chemical recovery section. For the Model A plant, it was specified in the model conditions (Appendix A) that the fans should be electrically driven, so that the electrical load is greater than that for the larger Model B plant.

The rich absorbent solution leaving the bottom of the absorbing tower is pumped to the chemical recovery section, where the sulfur dioxide is released from the bisulfite solution and the sulfite is reformed. A mixture of the sulfur dioxide and steam leaves the system and goes first to the main condenser, which is operated at about 200°F. Most of the steam is condensed, but the temperature is high enough that little of the sulfur dioxide is absorbed in the condensate, which can be returned to the system as makeup water. The hot, wet sulfur dioxide stream leaving the receiving tank of the main condenser enters the second condenser, where it is cooled to about 120°F. Most of the remaining water vapor condenses, but the condensate contains a high concentration of dissolved sulfur dioxide. This second condensate stream must therefore be fed to the condensate stripper, where the sulfur dioxide is recovered by stripping with a small amount of steam. The overhead from the stripper joins the main gas stream from the second condenser receiving tank. At this point the main stream of sulfur dioxide gas is at about 120°F and saturated with water vapor, which then constitutes less than 12 percent

by volume of the gas stream. The stripper overhead contains a much higher concentration of water vapor, but is only a small fraction of the main gas stream. The wet product sulfur dioxide is heated to prevent subsequent condensation of water vapor, then compressed to about 8 psig for delivery to an adjacent plant for subsequent manufacture of sulfuric acid or reduction to elemental sulfur.

The fly ash that is collected in the absorber must be purged from the system. Hence, a sidestream of the lean absorbent solution returning to the absorber is diverted through a centrifuge where the fly ash is removed and the clarified solution is returned to the absorber circuit. The amount of fly ash to be removed at this point and, hence, the size of the sidestream and centrifuge, is obviously dependent upon the efficiency of the gas cleaning system ahead of the absorber.

To prevent the buildup of an excessive sodium sulfate concentration in the absorbent, a purge stream of the lean absorbent solution is withdrawn. In the original design supplied by Wellman-Lord for this study, no provision was made for recovery of the sodium base from the absorbent purge stream because it was believed that the amount of sulfite (or sulfur dioxide) oxidized to sulfate would be very low. The author believes that the amount of sulfur dioxide oxidized will probably be much higher and that recovery of the purged base will be an economic necessity. If the amount of sulfur dioxide oxidized does not exceed perhaps 1 to 2 percent of that absorbed, it may be feasible to recausticize the sodium sulfate-sodium sulfite mixture with lime, precipitating the sulfate and sulfite. The recovered sodium hydroxide solution may have to be carbonated to precipitate calcium ions remaining in solution. If calcium ions remain in the solution, they may eventually produce scaling problems in the recovery system. As a guesstimate, the capitalized cost of operating such a recovery system might be about $12 to $15 per ton of sodium hydroxide recovered, including the cost of lime, makeup sodium hydroxide, operation, and depreciation of equipment. The higher figure, $15/ton, is assumed in this study.

In the purging of sulfate, a portion of the sulfur dioxide is lost because of purging of the associated sulfite. The author calculated the sulfur dioxide loss as 9.5 percent of that absorbed from the flue gas, assuming that 1.0 percent is oxidized and that 40 percent of the sulfur trioxide in the flue gas is collected in the absorber to form additional sulfate. If the amount of sulfur dioxide oxidized should be as much as perhaps 4 to 6 percent of that absorbed, the sodium hydroxide could still be recovered by the recausticization process, but the amount of sulfur dioxide lost in purging would be a substantial fraction of the total. In such case, there is a need for a system for separating and purging the sodium sulfate that will not result in loss of the sulfite.

VI PROCESS DATA AND COST ESTIMATES

A. Material Flows

The estimated quantities of coal burned in the two hypothetical power plants -- and hence the calculated quantities of flue gas, sulfur compounds, and fly ash to be handled -- were fixed by the assumed heat rates (see Appendix A). The heat rates assumed for the Model A and Model B power plants were 10,000 and 9,600 Btu/kwh, respectively. These values are probably between 5 and 10 percent higher than those that might reasonably be obtained at comparable actual plants. Although the assumed heat rates are perhaps excessively conservative, they are well within the probable precision of the other estimates made for the study.

Although the calculated volume of flue gas generated was increased by use of the high values of heat rate, no allowance was made for any increase in gas volume resulting from air infiltration into the flue gas handling system, which occurs in actual practice.

The estimated rates and conditions of gas flow through the Model A and Model B Wellman-Lord absorbing sections are presented in Table V. The treated flue gas leaving the absorber will have a water vapor content somewhat below the saturation value because of the vapor pressure lowering by the salts in the absorbent solution. The author assumed that there is no change in the moisture content of the gas after it leaves the prescrubber, saturated at 128°F.

The material balances for sulfur, fly ash, and absorbent chemicals are presented in Tables VI and VII, which include the sulfur balances for the alternative sulfur dioxide conversion plants (contact sulfuric acid and sulfur dioxide reduction). The performance of the prescrubber and absorber on sulfur trioxide and fly ash was based largely on estimates or assumptions by the author. The sulfur trioxide emitted to the atmosphere would actually be in the form of sulfuric acid mist. The original estimates of sodium hydroxide entrainment loss provided by

Wellman-Lord were tenfold higher than the values adopted by the author, who believes that the values given in Tables VI and VII should be reasonable unless reentrainment should take place in the eliminator.

Table V

FLUE GAS FLOWS IN WELLMAN-LORD
SYSTEM ABSORPTION SECTIONS

Gas Stream	Temp (°F)	Water Vapor Content (%)	Gas Flow Rate (ACFM)	
			Model A	Model B
Entering Prescrubber	325	8.08	1,564,360	3,003,230
Leaving Absorber	130	14.4	1,262,710	2,424,140

B. Capital Cost Estimates

The estimates of capital investment for the sulfur dioxide recovery plants presented in Table VIII, IX, and XII were prepared by Wellman-Lord with three exceptions. The author added allowances to each plant model for ductwork connecting the outlets of the absorbers to the stacks. The original estimates did not clearly indicate inclusion of such estimates.

The author also provided the credit for the stack for the Model B plant, which represents the difference between the cost of an 800-foot stack used with no sulfur oxide emission controls, and that of a 500-foot stack used when emission controls are applied (see Appendix C). The same stack credit was assigned arbitrarily to both the Cat-Ox and Wellman-Lord control systems for the Model B plant. The precision of the estimate is low, and the credit is provided mostly for purposes of illustration. As is discussed in Appendix C, some reduction of stack height in this model case appears justified. It appears to the author, as to some others,[4] that the problems of securing dispersion of scrubbed stack gases have been overstressed.

The estimates of capital investment for the contact sulfuric acid plants (Tables X and XIII) were taken from the curve of Fig. D-1, Appendix D of Part II of this report. The curve was derived from the data of Connor,[6] and gives only the cost of the acid-making section of a contact plant; since the feed to the plant consists of concentrated sulfur dioxide, no gas-producing section is required. The plants were sized for the total amount of sulfur dioxide absorbed by the Wellman-Lord systems without allowance for losses.

The estimates of capital investment for the sulfur dioxide reduction plants (Tables XI and XIV) were based on an estimate made for a conceptual design produced by Allied Chemical Corporation.[28] The specific values were taken from the curve of Fig. E-1, Appendix E of Part II of this report. The plants were sized for the total amount of sulfur dioxide absorbed by the Wellman-Lord systems, without allowance for losses.

C. Operating Cost Estimates

The estimates of electrical power, steam, water, and labor were developed by Wellman-Lord (Tables IX and XIII), and the corresponding operating costs were calculated by application of the appropriate cost factors. The ratio of sodium hydroxide and hydroquinone makeup to sodium sulfate purged was supplied by Wellman-Lord, but the author made his own estimates of the amount of sulfate formed, as is discussed above in Section V.

Wellman-Lord did not provide an estimate of annual maintenance cost, but the author adopted the figure of 3 percent of capital investment used generally for other systems in this study.

The operating requirements for the contact sulfuric acid plants were adopted from ranges of values presented in the literature.[6,8] Those for the sulfur dioxide reduction plants were taken from the report by Allied Chemical Corporation.[28]

Table VI

POWER PLANT MODEL A
MATERIAL BALANCES FOR SO_2 RECOVERY SYSTEM

Item	Quantity (lb/hr)	Percent of Input
I. Wellman-Lord System		
A. Sulfur Dioxide		
1. Input	22,200	100
2. Absorbed	19,980	90
3. Emitted	2,220	10
4. Oxidation and purge loss	1,893	8.52
5. Recovered	18,090	81.48
B. Sulfur Trioxide		
1. Input	391.0	100
2. Collected in prescrubber	195.5	50
3. Collected in absorber	156.4	40
4. Emitted	39.1	10
C. Fly Ash		
1. Input	1,405	100
2. Collected in prescrubber	1,265	90
3. Collected in absorber	134	9.5
4. Emitted	7.0	0.5
D. Sodium Hydroxide Losses		
1. Entrainment	12.6	
2. Reaction with SO_3	971	
3. Oxidation of SO_2	1,552	
E. Hydroquinone Loss	4.6	

(continued)

Table VI (Concluded)

Item	Quantity (lb/hr)	Percent of Input
II. Sulfuric Acid Plant		
A. Sulfur Dioxide		
1. Input	18,090	100
2. Converted to acid	17,730	98
3. Emitted	362	2
B. Sulfuric Acid Produced	27,150	
III. Sulfur Dioxide Reduction Plant		
A. Sulfur Dioxide		
1. Input	18,090	100
2. Converted to sulfur	16,760	92.63
3. Emitted	1,330	7.37
B. Sulfur Produced	8,380	
IV. W-L System plus Acid Plant		
A. Sulfur Dioxide		
1. Emitted	2,582	11.63
2. Usefully recovered	17,730	79.85
3. Lost in purge	1,893	8.52
V. W-L System plus Reduction Plant		
A. Sulfur Dioxide		
1. Emitted	3,550	16.00
2. Usefully recovered	16,760	75.48
3. Lost in purge	1,893	8.52

Table VII

POWER PLANT MODEL B
MATERIAL BALANCES FOR SO_2 RECOVERY SYSTEM

Item	Quantity (lb/hr)	Percent of Input
I. Wellman-Lord System		
A. Sulfur Dioxide		
1. Input	42,619	100
2. Absorbed	38,360	90
3. Emitted	4,262	10
4. Oxidation and Purge Loss	3,634	8.52
5. Recovered	34,720	81.48
B. Sulfur Trioxide		
1. Input	750.7	100
2. Collected in prescrubber	375.3	50
3. Collected in absorber	300.3	40
4. Emitted	75.1	10
C. Fly Ash		
1. Input	269.8	100
2. Collected in prescrubber	242.8	90
3. Collected in absorber	25.6	9.5
4. Emitted	1.35	0.5
D. Sodium Hydroxide Losses		
1. Entrainment	24.2	
2. Reaction with SO_3	1,864	
3. Oxidation of SO_2	2,980	
E. Hydroquinone Loss	8.8	

(continued)

Table VII (concluded)

Item	Quantity (lb/hr)	Percent of Input
II. Sulfuric Acid Plant		
A. Sulfur Dioxide		
1. Input	34,720	100
2. Converted to Acid	34,030	98
3. Emitted	694	2
B. Sulfuric Acid Produced	52,100	
III. Sulfur Dioxide Reduction Plant		
A. Sulfur Dioxide		
1. Input	34,720	100
2. Converted to Sulfur	32,160	92.63
3. Emitted	2,559	7.37
B. Sulfur Produced	16,080	
IV. W-L System plus Acid Plant		
A. Sulfur Dioxide		
1. Emitted	4,956	11.63
2. Usefully Recovered	34,030	79.85
3. Lost in Purge	3,634	8.52
V. W-L System plus Reduction Plant		
A. Sulfur Dioxide		
1. Emitted	6,281	16.00
2. Usefully Recovered	32,160	75.48
3. Lost in Purge	3,634	8.52

Table VIII

CAPITAL INVESTMENTS FOR WELLMAN-LORD
SO_2 RECOVERY PLANTS[1]

Model A	
Absorber System	$ 3,640,000
Chemical Recovery System	3,600,000
Total	$ 7,240,000
	$14.50/kw
Model B	
Absorber System	$ 7,460,000
Chemical Recovery System	5,400,000
Total	$12,860,000
Credit for Stack Height Reduction	850,000
Net total	$12,010,000
	$12.00/kw

[1] Not including auxiliary plants for conversion of concentrated sulfur dioxide. Capital investment does not include land, spares, interest on investment during construction, start-up expense, working capital, or royalty on process.

Table IX

POWER PLANT MODEL A
SUMMARY OF CAPITAL AND OPERATING COSTS OF WELLMAN-LORD SYSTEM

Item	Quantity	Cost Basis	Cost
Capital Investment		$	7,240,000
Annual Costs			
A. Fixed Charges		14.5% of capital investment	1,049,800
B. Direct Operating Costs			
1. Operating labor	25,667 hr/yr	$3.75/hr	96,250
2. Supervision	7,000 hr/yr	$4.75/hr	33,250
3. Payroll benefits		25% of labor + supervision	32,375
4. Maintenance		3% of capital investment	217,200
5. Electricity	8,500 kw	0.55¢/kwh	327,400
6. Steam	140,000 lb/hr	60.2¢/1000 lb	589,960
7. Scrubbing water	550 gpm	2¢/1000 gal	4,620
8. Process water	14,000 lb/hr	4.8¢/1000 lb	4,700
9. Cooling water	5,500 gpm	2¢/1000 gal	46,200
10. Sodium hydroxide--makeup	12.6 lb/hr	$65/ton	2,870
11. Sodium hydroxide--recovered	2,523 lb/hr	$15/ton	132,460
12. Hydroquinone	4.6 lb/hr	82¢/lb	26,400
C. Indirect Costs			
1. Overhead		50% of labor + supervision + maintenance	173,350
Total Annual Cost (Gross)		$/yr	2,736,835
		Mills/kwh	0.782
		¢/million Btu	7.82
		$/ton of coal	2.08

Table X

POWER PLANT MODEL A
SUMMARY OF CAPITAL AND OPERATING COSTS FOR CONTACT SULFURIC ACID PLANT

Item	Quantity	Cost Basis	Cost
Capital Investment		$	790,000
Annual Costs			
A. Fixed Charges		14.5% of capital investment	114,550
B. Direct Operating Costs			
1. Operating labor	7,000 hrs/yr	$3.75/hr	26,250
2. Supervision	1,750 hrs/yr	$4.75/hr	8,310
3. Payroll benefits		25% of labor + supervision	8,640
4. Maintenance		3% of capital investment	23,700
5. Electricity	25 kwh/ton acid	0.55¢/kwh	13,070
6. Process water	20 gal/ton acid	20¢/1,000 gal	380
7. Cooling water	7,000 gal/ton acid	2¢/1,000 gal	13,300
C. Indirect Costs			
1. Overhead		50% of labor + supervision + maintenance	29,130
Total Annual Cost (Gross)		$/yr	237,330
		Mills/kwh	0.068
		¢/million Btu	0.68
		$/ton coal	0.18

Table XI

POWER PLANT MODEL A
SUMMARY OF CAPITAL AND OPERATING COSTS FOR SULFUR DIOXIDE REDUCTION PLANT

Item	Quantity	Cost Basis	Cost
Capital Investment		$	1,040,000
Annual Costs			
A. Fixed Charges		14.5% of capital investment	150,800
B. Direct Operating Costs			
1. Operating labor	54 man-hr/day	$3.75/hr	59,060
2. Supervision	8 man-hr/day	$4.75/hr	11,080
3. Payroll benefits		25% of labor + supervision	17,540
4. Maintenance		3% of capital investment	31,200
5. Electricity	161 kw	0.55¢/kwh	6,200
6. Methane	53,200 CF/hr	30¢/1,000 CF	111,720
7. Fuel gas	3,080 CF/hr	30¢/1,000 CF	6,470
8. Cooling water	7,600 gal/hr	2¢/1,000 gal	1,060
9. Boiler feed water	90 gal/hr	40¢/1,000 gal	250
10. Catalyst	22,800 lb/yr	13¢/lb	2,960
11. (Low pressure steam)	(6,690 lb/hr)	(25¢/1,000 lb)	(11,710)*
C. Indirect Costs			
1. Overhead		50% of labor + supervision + maintenance	50,670
Total Annual Cost (Gross)		$/yr	437,300
		Mills/kwh	0.125
		¢/million Btu	1.25
		$/ton of coal	0.33

* Credit.

Table XII

POWER PLANT MODEL B
SUMMARY OF CAPITAL AND OPERATING COSTS FOR WELLMAN-LORD SYSTEM

Item	Quantity	Cost Basis	Cost
Capital Investment		$	12,010,000
Annual Costs			
A. Fixed Charges		14.5% of capital investment	1,741,450
B. Direct Operating Costs			
1. Operating labor	25,667 hr/yr	$3.75/hr	96,250
2. Supervision	7,000 hr/yr	$4.75/hr	33,250
3. Payroll benefits		25% of labor + supervision	32,375
4. Maintenance		3% of capital investment	360,300
5. Electricity	4,960 kw	0.55¢/kwh	191,000
6. Steam	270,000 lb/hr	60.2¢/1000 lb	1,137,780
7. Scrubbing water	1,100 gpm	2¢/1000 gal	9,240
8. Process water	27,000 lb/hr	4.8¢/1000 lb	9,072
9. Cooling water	10,000 gpm	2¢/1000 gal	84,000
10. Sodium hydroxide--makeup	24.2 lb/hr	$65/ton	5,505
11. Sodium hydroxide--recovered	4,844 lb/hr	$15/ton	254,310
12. Hydroquinone	8.8 lb/hr	82¢/lb	50,510
C. Indirect Costs			
1. Overhead		50% of labor + supervision + maintenance	244,900
Total Annual Cost (Gross)		$/yr	4,249,942
		Mills/kwh	0.607
		¢/million Btu	6.32
		$/ton of coal	1.69

Table XIII

POWER PLANT MODEL B
SUMMARY OF CAPITAL AND OPERATING COSTS FOR CONTACT SULFURIC ACID PLANT

Item	Quantity	Cost Basis	Cost
Capital Investment		$	1,200,000
Annual Costs			
A. Fixed Charges		14.5% of capital investment	174,000
B. Direct Operating Costs			
1. Operating labor	7,000 hr/yr	$3.75/hr	26,250
2. Supervision	1,750 hr/yr	$4.75/hr	8,310
3. Payroll benefits		25% of labor + supervision	8,640
4. Maintenance		3% of capital investment	36,000
5. Electricity	25 kwh/ton acid	0.55¢/kwh	25,070
6. Process water	20 gal/ton acid	20¢/1,000 gal	730
7. Cooling water	7,000 gal/ton acid	2¢/1,000 gal	25,530
C. Indirect Costs			
1. Overhead		50% of labor + supervision + maintenance	35,280
Total Annual Cost (Gross)		$/yr	339,810
		Mills/kwh	0.049
		¢/million Btu	0.50
		$/ton of coal	0.13

Table XIV

POWER PLANT MODEL B
SUMMARY OF CAPITAL AND OPERATING COSTS FOR SULFUR DIOXIDE REDUCTION PLANT

Item	Quantity	Cost Basis	Cost
Capital Investment		$	1,520,000
Annual Costs			
A. Fixed Charges		14.5% of capital investment	220,400
B. Direct Operating Costs			
1. Operating labor	54 man-hr/day	$3.75/hr	59,060
2. Supervision	8 man-hr/day	$4.75/hr	11,080
3. Payroll benefits		25% of labor + supervision	17,540
4. Maintenance		3% of capital investment	45,600
5. Electricity	309 kw	0.55¢/kwh	11,900
6. Methane	102,100 CF/hr	30¢/1,000 CF	214,410
7. Fuel gas	5,900 CF/hr	30¢/1,000 CF	12,390
8. Cooling water	14,580 gal/hr	2¢/1,000 gal	2,040
9. Boiler feed water	174 gal/hr	40¢/1,000 gal	490
10. Catalyst	43,800 lb/yr	13¢/lb	5,700
11. (Low pressure steam)	(12,800 lb/hr)	(25¢/1,000 lb)	(22,400)*
C. Indirect Costs			
1. Overhead		50% of labor + supervision + maintenance	57,870
Total Annual Cost (Gross)		$/yr	636,080
		Mills/kwh	0.091
		¢/million Btu	0.95
		$/ton of coal	0.25

* Credit.

VII GENERAL DISCUSSION

A. Evaluation of the Wellman-Lord SO_2 Recovery System

The Wellman-Lord process using sodium-base absorbent, described above, appears to be technically feasible. The regeneration of the absorbent, which is, in the writer's opinion, the only really novel aspect of the process, appears to be feasible and to offer a significant reduction of the steam consumption from the levels required by direct steam stripping of the rich absorbent. The capital investment in the Wellman-Lord chemical recovery system is probably much higher than that for a simple steam-stripping system. However, the associated fixed charges on the higher investment should be more than compensated by the reduction of twofold or threefold in the steam requirement.

The use of the potassium-base absorbent should also be technically feasible, but the most recent information available indicates that the economic advantages originally anticipated by Wellman-Lord are not realizable.

Serious operating problems were encountered in the demonstration plant using the potassium-base absorbent. Most of these were of a mechanical nature and not related to inherent defects in the basic absorption system. However, they had a critical effect on the testing program. The most serious problem was evidently the plugging of the prescrubber (and sometimes parts of the absorber system) by fly ash. The prescrubber was defective in mechanical design and inadequate in efficiency. However, the prescrubber proposed by Wellman-Lord for use in the Model A and B power plant control systems is an established design that has been widely used, and it should provide higher efficiency as well as fewer problems with plugging. Even should this prescrubber design not prove entirely satisfactory from the standpoint of plugging, there are other designs available that should be superior in this respect and should deliver the same efficiency at the same pressure drop.[21,22]

These alternative scrubber designs should also be comparable in cost. Before the final design of prescrubber is chosen for a large recovery plant (commercial or demonstration), operating tests should be made with small pilot plant units.

The use of the high-efficiency electrostatic precipitators to collect the bulk of the fly ash is probably desirable, if not a practical necessity, so long as only a low-energy type of prescrubber is employed. If the precipitator should be omitted entirely or should have an efficiency of no more than 90 percent, the prescrubber should be of the high-energy, high-efficiency type. The reported performance of scrubbers on fly ash from boilers firing pulverized coal[19,20] indicates that the fly ash is actually relatively easy to collect, and that high collection efficiencies can be attained with moderate power inputs (6 to 10 inches of water pressure drop).

In the demonstration plant some plugging problems were encountered at the bottom of the absorption tower because of the formation of potassium pyrosulfite crystals. The problem arose because the pyrosulfite is the less soluble of the two potassium salts and reached its highest concentration at the bottom of the tower. In the sodium-base system, where the pyrosulfite is the more soluble of the two salts, such a problem is more easily avoided unless excessive evaporation of water from the solution by the flue gas should take place.

The type of absorber proposed by Wellman-Lord is neither the only, nor necessarily the best, that might be employed. Stone & Webster Engineering Corp.[24] prepared estimates of the costs of various types of absorbers for use with the same sodium sulfite-bisulfite system. The estimated costs of the absorber systems ranged from $2.25 to $4.76/ACFM, based on exit gas conditions. The estimated costs of the Wellman-Lord absorber systems for Model A and Model B are $2.88 and $2.08/ACFM, respectively. Except for the factor of cost, however, Stone & Webster was unable to present any information that provides a clear choice between absorber designs.

A major unresolved question concerning the Wellman-Lord system is the amount of oxidation of the sulfite to sulfate. Some data should be acquired during 1970 from the commercial plant being constructed for use on sulfuric acid plant tail gas, but they are unlikely to be generally applicable even when they do become available. The general effectiveness of oxidation inhibitors has not been established. There appears to be no way short of actual experience under the proposed operating conditions by which the amount of sulfite oxidation can be predicted with much confidence. Experience with other recovery systems[2,5,9,14,18] suggests that the amount of the sulfur dioxide oxidized might be in the range of 0.2 to 2.0 percent, but could possibly be as much as 5 to 10 percent or even higher, dependent upon the conditions of the particular case. Two of the most critical factors are the oxygen concentration in the gas[2,18] and the presence of oxidation catalysts in the absorbent solution.[10,11,23]

High efficiency cleaning of the flue gas ahead of the absorber will be helpful because it will reduce the introduction of potential oxidation catalysts into the absorbent. However, some of the catalysts are effective at very low concentrations.[10,11]

As is discussed above in Section V, oxidation of as little as 4 to 6 percent of the sulfur dioxide can result in loss of a substantial fraction of the total sulfur dioxide during purging of the sulfate. Therefore, there is potentially a need for an alternative method for separating the sodium sulfate from the sodium sulfite so that most of the latter can be returned as such to the absorption cycle.

B. By-Product Values

Over the period of about one year (1969), sulfur supplies have passed abruptly from shortage to surplus, and it is now difficult if not impossible to determine whether there is an established price for sulfur. The Gulf Coast price of sulfur produced by the Frasch process has long been the base level for world sulfur prices, but it is apparently not so at present, and it is uncertain whether it will be again. In the previous study of the nonferrous smelting industry[15] it was estimated that the average Gulf Coast price of sulfur over the period

to 1975 might be $30/long ton. When the estimate was made, the Gulf Coast price was over $40/long ton; currently sulfur is reported to be selling in some areas for less than $15/long ton. Although sulfur prices could rise again during the next five years, the figure of $30 now appears to be at the optimistic end of the probable range of prices that may prevail to 1975 (see Appendix D). Nevertheless, it has been used as the basis for by-product price estimates in this study because there appears to be none that can be used with significantly greater confidence, and the results still have some usefulness if their limitations are understood.

Even the original estimate of $30/long ton f.o.b. the Gulf Coast was based on the assumption that no very substantial quantity of sulfur would be recovered at power plants before 1975. The amounts of sulfur and sulfur by-products potentially recoverable from the coal consumed by power plants (whether from the coal itself or from the flue gases) are so large that recovery of a substantial part of them could completely disrupt existing local sulfur markets. The estimates made of sulfur and sulfuric acid prices that might be obtained by the Model A and Model B power plants (see Appendix D) embody the tacit assumption that no other large power plants near the assumed locations will also be recovering sulfur by-products.

In summary, it appears that the credits assigned to sulfur by-products from the Model A and Model B plants are the highest that can be reasonably expected in the period up to 1975. The actual credits might be much lower, and the situation that may exist after 1975 is unpredictable on the basis of currently available information. Conservatively, the long-term assessment of the economics of recovery processes should probably be based on the gross costs without allowance for by-product credits.

It is sometimes suggested that recovery processes can be more economically applied to power plant flue gases if coals with extra high sulfur contents are used. Since the sulfur dioxide concentration in the flue gas will be increased, the unit cost of production of the sulfur

by-product should indeed be reduced. However, if the by-products must enter markets that are already glutted, the additional quantity of material may merely exert an increased downward pressure on prices. The actual advantage, if any, of using the high sulfur coal would probably be related to lower purchase prices for the coal.

C. Variations of Bases of Cost Estimates

The choice of some of the economic factors assumed in making the cost estimates was to a considerable degree arbitrary, as it has been in similar studies made by other workers. There is no general agreement among different workers on what constitutes a realistic set of bases. The result is that estimates for different control systems have been made on a wide variety of bases -- frequently only vaguely defined -- so that the results cannot be directly compared. In this study, it has been assumed that valid comparison of different control systems is more important than absolute accuracy, and hence that it is more important that the bases used should be uniform than that they should be the "correct" ones -- whatever the latter might be. An effort has been made to describe fully all bases used, so that the results can be recalculated to any different bases that other persons may wish to use. Although the bases chosen for the models (Appendixes A and B) are believed to be generally reasonable, it must be noted that other possible assumptions regarding some factors might be equally reasonable but lead to markedly different cost estimates.

1. Load Factor

The assumed annual load factor of 80 percent (equivalent to 7,000 hours per year of operation at full load) was specified by DPCE-NAPCA. It appears to be fairly reasonable for large, new, or relatively new baseload plants such as those typified by Models A and B. It is less certain whether the 80 percent factor can be assumed to hold over the whole of the assumed life (20 years) of the emission control systems. The Federal Power Commission assumed (Hydroelectric Power Evaluation, Report P-35, March 1968) that the annual load factor will be relatively high over the first half of the estimated 30- to 35-year service life

of a power plant, then will diminish, giving a lifetime average of 55 to 60 percent. Dennis and Bernstein[7] assumed a load factor of 60 percent over an assumed 11-year life for the emission control systems. Although their assumptions may be somewhat too conservative, the author believes that the average factor of 80 percent assumed in the present study is probably too optimistic.

Because of the high fixed charges associated with the emission control systems, the load factor exerts great leverage in determining the incremental cost of producing electricity. The estimates of incremental unit costs presented in Tables I through IV are probably as low as they can be expected to be with respect to load factor. The actual costs would probably be higher.

2. Amortization Period

Depreciation for the emission control systems was based on an assumed useful life of 20 years. It was considered that any control system that might go into service at the present time will be obsolete within 20 years, if not in a shorter period. It was assumed that tax laws and the regulations of public utility commissions will permit the use of this period in determining taxes and rates for electrical power. At this time there appear to be no precedents to indicate how the existing depreciation schedules of the tax laws might be interpreted in some specific cases. Dennis and Bernstein[7] assumed a depreciation period of 11 years, as specified for chemical plants by the Internal Revenue Service. Their choice of depreciation period apparently was influenced by a conviction that the recovery processes would become obsolete within the 11-year period as well as by the IRS regulations.

Some components of sulfur dioxide recovery systems are relatively conventional items of chemical plant equipment. For example, a basically standard contact process plant would be used to convert concentrated sulfur dioxide from a recovery system to sulfuric acid. Allowances for depreciation of the acid plant might then depend upon whether the plant was owned by the power company and considered to be part of the utility, or owned by an adjacent chemical company and considered to be part of a chemical plant.

For the purposes of the present study, such considerations were ignored and it was assumed that all components of the control and recovery systems would be depreciated over the 20-year period. Nevertheless, it must be recognized that the assigned useful life of the control system, like the load factor of the power plant, can exert a very great weight in determining the incremental unit cost of producing electricity. The author believes that, on the whole, the assumption of a useful life of 20 years is as optimistic as can now be made for any control system, and that the actual useful lifetimes might well prove to be shorter. In such case, the incremental unit costs of producing electricity given in Tables I through IV could be substantially increased.

3. Fixed Charges

The total fixed charge of 14.5 percent of the capital investment per year was adopted from a report by the Tennessee Valley Authority,[25] which was in turn adapted from guidelines suggested by the Federal Power Commission, and embodies the assumption of a 20-year depreciation period. Recent increases in both taxes and the cost of capital are producing rises in the fixed charges associated with electric power generation. The figure of 14.5 percent is therefore lower than may be appropriate at the present time. It has been used in the present study primarily for the sake of consistency with previous studies, and because the uncertainties in other factors assumed appear to make greater elaboration of the capital charge estimates of limited utility. The trend in taxes and capital costs will, however, tend to increase the incremental unit costs of electrical power generation estimated in Tables I through IV.

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Appendix A

MODELS FOR HYPOTHETICAL POWER PLANTS

For the hypothetical power plant models, DPCE-NAPCA specified the following conditions:

1. The two model plants shall be an existing 500-MW plant located in central Pennsylvania and a new 1000-MW plant located in the Midwest on a large navigable river.

2. Both plants shall have dry-bottom boiler furnaces and shall use coal containing 3 percent of sulfur and 9 percent of ash. The flue gases shall contain 0.21 percent of sulfur oxides.

3. The base load shall be 7,000 hours per year (load factor 80 percent).

4. The existing plant shall have an economizer with an exit gas temperature of 750°F and an electrostatic precipitator operating at 325°F. The new plant shall have no temperature constraints on gas cleaning equipment. The height of the stack for the existing plant shall be 500 feet. Sufficient space shall be available to permit installation of any equipment needed.

5. The efficiency of removal of the sulfur dioxide shall be 90 percent. Regional air and water pollution regulations shall be applicable.

DPCE-NAPCA also supplied an analysis of the coal assumed to be burned and a calculated analysis of the flue gas formed:

Analysis of Coal, As Fired	
Constituent	lb/100 lb, Coal as Fired
C (burned)	72.7
H_2	5.1
N_2	1.4
S	3.0
O_2	6.9
H_2O	1.0
Ash	9.0
	100.0

Flue Gas Analysis[1]	
Component	Percent by Volume
CO_2	13.89
SO_2 and SO_3	0.2148
O_2	3.301
N_2	74.514
H_2O	8.08
	100.0

[1] Excess air used = 20 percent. Moisture content of air = 0.013 lb/lb dry air.

Quantity of Flue Gas Formed

Basis	Lb Moles Gas/100 lb Coal Burned
Wet	43.67
Dry	40.14

The remainder of the model conditions were derived by SRI from the foregoing conditions, supplemented by assumptions based on data in the literature, or on the experience or judgment of the author.

A. Conditions Common to Both Model Plants

1. Heating Value of Coal

   13,325 Btu/lb (calculated from DuLong Formula)

2. Flue Gas Generated

   Volume of gas per 100 lb of coal burned (wet basis)

   = 16,880 SCF (70°F, 1 atm)

   = 25,013 CF (325°F, 1 atm)

   = 43,335 CF (900°F, 1 atm)

3. Air Inleakage into Flue Gas Handling System

   Assumed nil

4. Fly Ash

   Ash = 9.0 lb/100 lb coal burned

   Unburned carbon = 0.35 lb/100 lb coal burned

   Total ash and unburned carbon = 9.35 lb/100 lb coal burned

   Fly ash (ash + carbon) leaving furnace in flue gas

   = 80 percent of total

   = 7.49 lb/100 lb coal burned

   Fly ash concentration in flue gas = 3.105 grains/SCF (70°F, 1 atm, wet basis)

5. Sulfur Oxides Concentrations

   $SO_2 + SO_3$ = 0.2148 percent by volume, wet basis

   SO_3 = 30 ppm = 0.003 percent by volume, wet basis

   SO_2 = 0.2118 percent by volume, wet basis

6. Sulfur Emission

All of sulfur assumed to go into flue gas

Total sulfur emitted = 3.0 lb/100 lb coal burned

Sulfur emitted as SO_3

= 0.0417 lb S/100 lb coal burned

= 0.1042 lb SO_3/100 lb coal burned

Sulfur emitted as SO_2

= 2.958 lb S/100 lb coal burned

= 5.916 lb SO_2/100 lb coal burned

7. Efficiency of Emission Control

Sulfur oxide removal efficiency = 90 percent

Allowable discharge of sulfur (as SO_2, SO_3, or H_2SO_4) not to exceed 10 percent of the input sulfur to the collection system, or 0.3 lb S/100 lb coal burned

8. Fuel Available for Reheating Flue Gas

No. 2 fuel oil (sulfur content 0.35 percent or lower)

9. Reductant Available for Reducing Sulfur Dioxide

Natural gas (heating value 1000 Btu/CF)

B. Model A Plant

1. Location

Altoona, Pennsylvania

2. Generating Capacity

500 MW

3. Temperature of Flue Gas

At exit from economizer: $750^{o}F$

At electrostatic precipitator: $325^{o}F$

4. Electrostatic Precipitator Efficiency

95 percent

5. Stack Height

   500 feet

6. Drives for Fans and Pumps

   Electric

7. Heat Rate

   10,000 Btu/kwh

8. Coal Consumption

   0.7505 lb/kwh

9. Coal Firing Rate

   375,250 lb/hr
   187.63 tons/hr

10. Flue Gas Flow Rate

    1,056,204 SCFM (70^{o}F, 1 atm, wet basis)
    1,564,355 CFM (325^{o}F, 1 atm, wet basis)

11. Fly Ash Leaving Furnace in Flue Gas

    28,106 lb/hr
    14.053 tons/hr

12. Fly Ash in Flue Gas Leaving Precipitator

    Concentration = 0.1553 grain/SCF (70^{o}F, 1 atm, wet basis)
    Mass emission rate = 1405.3 lb/hr

13. Sulfur Oxide Emissions

    SO_2 = 22,200 lb/hr
    = 11.100 tons/hr
    = 266.40 tons/day
    SO_3 = 391.01 lb/hr
    = 9384 lb/day

C. Model B Plant

1. Location

   Cairo, Illinois

2. Generating Capacity

   1000 MW

3. Gas Cleaning System

   No temperature limits apply. Unless some other gas cleaning system is specified as an integral part of the sulfur oxide control process, the power plant will be assumed to be equipped with an electrostatic precipitator operating on flue gas at a temperature of 325°F and having an efficiency of 99.5 percent.

4. Stack Height

   The plant will be assumed to have an 800-foot stack when not equipped with a sulfur dioxide control system, and a 500-foot stack when equipped with a sulfur dioxide control system of 90-percent efficiency.

5. Drives for Fans and Pumps

   Either electric or steam turbine

6. Heat Rate

   9600 Btu/kwh

7. Coal Consumption

   0.7204 lb/kwh

8. Coal Firing Rate

   720,400 lb/hr
   360.20 tons/hr

9. Flue Gas Flow Rate

2,027,690 SCFM(70°F, 1 atm, wet basis)

3,003,230 CFM (325°F, 1 atm, wet basis)

5,203,090 CFM (900°F, 1 atm, wet basis)

10. Fly Ash Leaving Furnace in Flue Gas

53,958 lb/hr

26.979 tons/hr

11. Fly Ash in Flue Gas Leaving Precipitator of 99.5 Percent Efficiency

Concentration = 0.01553 grain/SCF (70°F, 1 atm, wet basis)

Mass emission rate = 269.79 lb/hr

12. Sulfur Oxide Emissions

SO_2 = 42,619 lb/hr

= 21.309 tons/hr

= 511.43 tons/day

SO_3 = 750.66 lb/hr

= 18,016 lb/day

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Appendix B

COST FACTORS USED IN MODEL STUDIES

In an actual case, the busbar cost of electricity and the cost of steam generated at a power plant are, of course, related to the cost of fuel and other plant operating costs as well as to the fixed costs, which are in turn related to the capital investment and the plant load factor. In practice, it appears that the complexity of cost accounting makes it difficult for even a utility company to assign costs precisely for an individual power plant in its system.

In the present study, reasonable costs were assumed for individual items such as fuel, steam, and electrical power. No attempt was made to reconcile these costs, but it was appreciated that the values adopted may actually not be entirely consistent with one another.

A. Fuel Costs

Coal

The cost of coal was estimated from data in "Steam-Electric Plant Factors (1967)," published by the National Coal Association. The costs (¢/million Btu) for coal as burned in Pennsylvania power plants were averaged for coals having heating values greater than 13,000 Btu/lb, which had higher unit costs than the lower-grade coals. The average cost per million Btu was applied to the hypothetical coal assumed in the present study (heating value 13,325 Btu/lb) to give the following costs:

31.2¢/million Btu

$8.30/ton

The average costs of coal used in Illinois, Indiana, Ohio, and Pennsylvania were lower, but applied to coals having heating values in the range of 10,000 to 12,000 Btu/lb. Presumably these coals had generally higher ash contents than the hypothetical coal.

Natural Gas

An estimate of the cost of natural gas was taken as the rounded average of the average costs of natural gas burned at power plants in Illinois, Indiana, Ohio, and Pennsylvania. The data were taken from "Steam-Electric Plant Factors (1967)." The value adopted was:

30¢/million Btu

The heating value of the gas was assumed to be 1,000 Btu/CF.

No. 2 Fuel Oil

The cost of No. 2 fuel oil in central Pennsylvania was estimated at SRI to be:

$4.25/barrel

B. Water

The costs of water for various uses were taken from lists of standard factors drawn from the literature or were figures used for estimating purposes at SRI.

	Cost
Raw makeup water	2¢/1000 gal
Once-through cooling water (temperature 80°F)	2¢/1000 gal
Process water	20¢/1000 gal
Boiler feed water	40¢/1000 gal
Retreatment of process steam condensates for boiler use	38¢/1000 gal

C. Steam

The cost of high-pressure steam was very roughly estimated at SRI as:

65¢/1000 lb (2400 psig, 1050°F)

The cost of the steam includes the cost of boiler feed water treatment.

It was assumed that low-pressure steam was available as needed, obtained by extraction from the power plant turbines. The value of low-pressure steam extracted from turbines, or exhausted from a backpressure turbine, was assumed to be proportional to the value of the high-pressure steam in the ratio of the enthalpies of the steam at the two conditions.

D. Electrical Power

The busbar cost of electrical power generated and used in the power plant itself (including the emission control system), after inspection of data published by the Federal Power Commission and after discussions with staff of the Pacific Gas & Electric Co., was assumed to be as follows:

Generating cost	3.2 mills/kwh
Fixed charges	2.3 mills/kwh
Total cost	5.5 mills/kwh

E. Salaries and Payroll Benefits

Salaries

Operating labor	$3.75/hour
Supervision	$4.75/hour

Payroll benefits

25 percent of labor and supervision

F. Maintenance

Taken as an appropriate percentage of capital investment to cover both labor and materials.

G. Indirect Costs

Overhead

50 percent of labor, supervision, and maintenance

H. Fixed Charges

The total fixed charge was adopted from the Tennessee Valley Authority's Conceptual Design and Cost Study No. 2 (1969) for National Air Pollution Control Administration, "Sulfur Oxide Removal from Power Plant Stack Gas--Use of Limestone in Wet-Scrubbing Process." The breakdown of fixed charges (adapted from Federal Power Commission guidelines) is given as follows:

	Annual Percent of Capital Investment
Depreciation (20-year life)	5.00
Insurance	0.25
Cost of capital (capital structure assumed to be 50% debt and 50% equity)	
Bonds at 6% of average undepreciated investment	1.50
Equity at 11% of average undepreciated investment	2.75
Taxes	
Federal	2.80
State	2.20
Total fixed charges	14.50

Appendix C

STACKS FOR USE ON CONTROLLED SULFUR DIOXIDE EMISSION SOURCES

Evidently, no clear policy has yet been developed with respect to choice of the heights of stacks to be used on power plants equipped with sulfur dioxide emission controls. The use of a high stack on an uncontrolled emission source reduces the maximum ground-level concentration of sulfur dioxide at the point where the plume reaches the ground, but it does not, of course, reduce the total emission of sulfur dioxide that may affect areas beyond the range of influence of the stack. If emission controls are used as well, not only is the maximum ground-level concentration further reduced, but the quantity of sulfur dioxide affecting the area beyond the range of influence of the stack is reduced also.

If, on the other hand, the use of emission controls is accompanied by a reduction of the stack height, the reduction of the maximum ground-level concentration of sulfur dioxide may be partly lost, depending upon the balance between the reduction in stack height and the reduction in emission. The point of maximum ground-level concentration will be brought closer to the emission source.

There is reluctance to make major reductions in stack height, even with installation of sulfur dioxide emission controls, because of concern with other contaminants that may still remain in the flue gas and with the possibility of failures of the emission control system.[C4]

If the sulfur dioxide emission control system reduces the temperature of the flue gas, the effect is to reduce the effective stack height because the density of the gas is increased and hence its buoyancy is reduced.[C3,C4] When the hot gas is scrubbed, the effect of gas cooling is counteracted to some degree by the lowering of the gas density due to the increase in the water vapor content of the gas stream.

Particular concern has been expressed about "negative buoyancy." If the plume emerging from the stack is saturated and contains water droplets, the subsequent evaporation of the drops in the atmosphere may cool the gas sufficiently to cause the plume to descend rapidly.[C1,C2] The presence of the droplets in the plume may directly produce such a descent by increasing the effective density of the gas in the plume. However, the author believes that the presence in stack gas of any such relatively gross quantities of liquid would most likely be due to entrainment from the scrubber, which can be prevented. It has been suggested[C1] that at least some of the reported instances of rapid descent of plumes of scrubbed gas may have resulted from downwash produced by adjacent structures. Plumes of scrubbed gas that is saturated but still substantially above ambient temperature have been commonly observed to rise and disperse in the same general manner as do warm, dry plumes. Careful observations of such plume behavior have been reported by Boyer et al.[C1]

The necessity for reheating scrubbed stack gases to ensure adequate dispersion is not entirely clear, at least if the efficiency of the scrubber is high and some minimum stack height is used. Using a hypothetical 200-MW power plant unit with a 300-foot stack as an example, TVA[C4] calculated the effects of cooling (by scrubbing) and of reheating the flue gas on the maximum ground-level concentration of sulfur dioxide, assuming varying scrubber efficiencies. The results indicated that if the scrubber had an efficiency of at least 90 percent and the stack height was at least 300 feet, failure to reheat the gas would have only a small effect upon the percentage reduction in the maximum ground-level concentration of sulfur dioxide achieved by scrubbing. Applying a commonly used rule of thumb that a stack should be at least 2.5 times the height of adjacent structures, a typical power plant should in any case use a stack at least 300 feet in height in order to avoid possible downwash of the gas.

In the present study, it was assumed that in the Model A plant, the treated flue gas would be discharged through the existing 500-foot stack.

In the case of the Model B plant, it was assumed that the plant would be equipped with an 800-foot stack if it had no sulfur dioxide emission controls, and with a 500-foot stack if it were equipped with an emission control system of 90 percent efficiency. Such a reduction in stack height does not appear to be unreasonable. In the light of present knowledge, a minimum stack height of 500 feet for the controlled emission source seems appropriate.

The Model B control systems (both Cat-Ox and Wellman-Lord) were each given credit for the differential between the costs of the 800- and 500-foot stacks. A rough estimate of the cost differential for stacks 30 feet in inside diameter ($850,000) was made from Fig. 6-1 of "Control Techniques for Sulfur Oxide Air Pollutants."[C2]

REFERENCES

C1. Boyer, A.E., F. B. Kaylor, T. V. Ward, and F. J. Gottlich, Atmospheric Dispersion of Saturated Stack Plumes, Des. Oper. Air Pollut. Contr., Pap. MECAR Symp. 1968 (Pub. 1969), 32-40.

C2. National Air Pollution Control Administration, Control Techniques for Sulfur Oxide Air Pollutants, Publication No. AP-52 (Jan. 1969)

C3. Scorer, R. S., "Air Pollution," Pergamon Press, New York (1968)

C4. Tennessee Valley Authority, Sulfur Oxide Removal from Power Plant Stack Gas--Use of Limestone in Wet-Scrubbing Process, Conceptual Design and Cost Study No. 2 for National Air Pollution Control Administration (1969)

Appendix D

ESTIMATION OF THE VALUES OF SULFUR BY-PRODUCTS*

As applied to power plant flue gases, the Monsanto Cat-Ox system can produce only 78-percent sulfuric acid. The Wellman-Lord system produces concentrated sulfur dioxide, which has relatively insignificant markets as such but can be converted to elemental sulfur or to either 93-percent or 98-percent sulfuric acid.

The estimates of the sulfur by-product prices at the two hypothetical power plant sites (Altoona, Pennsylvania for the Model A plant, and Cairo, Illinois for the Model B plant) were based on an assumed Gulf Coast price for sulfur of $30/long ton. This Gulf Coast price was estimated during the previous study[D1] as a likely average over the period to 1975. Currently, sulfur prices have become chaotic as sulfur supplies have shifted abruptly from shortages to surpluses. Apparently it is scarcely possible to define an "established" price for sulfur. The development of sulfur surpluses has followed a continuing period of low activity in the market for fertilizers, which provides the largest outlet for sulfur.

In the next five years, the withdrawal of marginal producers and a possible revival of the fertilizer market may cause sulfur prices to rise from their present lows. However, the figure of $30/long ton now appears to be an optimistic one. It perhaps represents the upper end of the probable range of Gulf Coast prices to be encountered in the period to 1975. It cannot even be assumed that the Gulf Coast sulfur price will continue to maintain its previous status as the base line for world sulfur prices. Nevertheless, if the above limitations are recognized, the assumption of the $30/long ton price is probably as good as any that can be made at this time. Prices at other locations can be estimated by adding to the Gulf Coast price the cost of transportation from the Gulf

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*This material was prepared by F. Alan Ferguson, Industrial Economist, SRI.

Coast. This procedure assumes, of course, that the incursion of the sulfur from a new source will not be large enough to disrupt the local markets.

Model A Plant - Altoona, Pennsylvania

The closest and most ready markets for sulfur by-products produced at the hypothetical power plant Model A would be located in the Pittsburgh area, about 120 miles away. The acid users in the Pittsburgh region currently consume over 500,000 tons of acid per year, a large portion of which is used to clean (pickle) iron and steel products and to produce ammonium sulfate from the ammonia in coke oven gas. Both of these uses could employ 78-percent sulfuric acid as well as 93-percent or 98-percent acid, and the total market is more than large enough to absorb all the acid that could be produced at the Altoona plant. The Pittsburgh area could also absorb elemental sulfur that might alternatively be produced at Altoona.

The prices that could be obtained for the various possible by-products producible at Altoona will depend upon a number of factors, a few of which can be allowed for at the present time. On the basis of a Gulf Coast sulfur price of $30/long ton, the delivered price of elemental sulfur in the Pittsburgh area will be about $36.50/long ton. Since the elemental sulfur now being sold in the Pittsburgh area is merchant sulfur, that produced at Altoona could probably capture about half the existing market if sold at the same price of $36.50/long ton. Since the cost of transportation from Altoona to Pittsburgh would be about $3.50/long ton, the f.o.b. price of the sulfur at Altoona might be $33/long ton.

Estimating the price of Altoona sulfuric acid is more difficult because some of the Pittsburgh area acid is made for captive consumption and the rest is made and sold as merchant acid. The prices of captive and merchant acid can differ by as much as a factor of two. The price of merchant acid will vary according to whether the acid is sold on long term contracts or when and as needed (spot sales). However, relatively little acid is sold at the higher spot prices. The highest price the plant at Altoona could expect to receive for its acid at Pittsburgh

would be the same as that which would return a reasonable profit to an acid producer operating a contact acid plant and using Gulf Coast sulfur as raw material. The cost of producing acid from sulfur costing $36.50/long ton would be about $13/short ton (100 percent acid basis). With the allowance of about 15 percent as a return on investment, the corresponding price of acid would be $15/short ton (100 percent acid basis). The corresponding f.o.b. prices at Altoona can be estimated by subtracting the transportation charges of $4/ton (100 percent acid basis) for 93-percent acid and $5/ton (100 percent acid basis) for 78-percent acid, giving:

93-percent acid: $12/ton of 100 percent acid

78-percent acid: $11/ton of 100 percent acid

If it should be necessary to displace existing captive producers to capture adequate markets, the Altoona acid might have to sell at Pittsburgh for about $2/ton less than the cost of production from sulfur, or $11/ton. The corresponding f.o.b. prices at Altoona would then be:

93-percent acid: $7/ton of 100-percent acid

78-percent acid: $6/ton of 100-percent acid

A detailed market analysis would be required to determine whether the price of the Altoona acid would be nearer the upper end ($11 to $12/ton) or the lower end ($6 to $7/ton) of the price range estimated.

Model B Plant - Cairo, Illinois

Finding markets for elemental sulfur produced at a plant at Cairo should not be difficult. In 1965, over 1.2 million tons of elemental sulfur were barged along the Ohio River and the upper Mississippi River. Assuming a Gulf Coast sulfur price of $30/long ton, sulfur produced along the rivers should sell for $30 plus the cost of barging sulfur from the Gulf Coast to the actual plant site. The cost of barging elemental sulfur from the Gulf to Cairo, Illinois is about $2.90/long ton. Hence, sulfur produced at the Cairo plant should sell for about $33/long ton.

If sulfuric acid, rather than elemental sulfur, should be produced, finding markets for large quantities of acid might be substantially more difficult. In 1966, only about 130,000 tons of acid were barged along the Ohio River and 30,000 tons along the upper Mississippi River. From the summary (Table D-1) of uses of sulfuric acid near three sites on these two rivers, it is apparent that the only end use large enough to consume the quantities of acid that could be supplied from the Cairo plant would be the production of phosphate fertilizers.

A recovery system producing sulfuric acid at Cairo might supply either existing phosphate fertilizer manufacturers or a new plant located at or near Cairo to take advantage of the local source of acid. Supplying such a new fertilizer manufacturer would probably yield the highest price for the acid, although a more detailed market analysis would be required to determine whether this would actually be the case. Nevertheless, the sulfuric acid would have to be priced to provide the new fertilizer producer with a competitive advantage over the existing suppliers. The existing producers of phosphate fertilizers that are capable of supplying fertilizer to the midwestern market at the lowest prices are probably those who manufacture their product at plant sites in Louisiana and ship it to the markets by barge. The price of the by-product acid at Cairo must therefore permit the new producer to make his fertilizer for less than the Louisiana producers can make and ship theirs.

In Table D-2 a comparison is made of the costs of producing 54-percent wet process phosphoric acid in Louisiana and at the hypothetical Cairo site. The data indicate that the Cairo producer could pay about $32.20 for the amount of 100-percent sulfuric acid necessary to make 1.85 tons of 54-percent (P_2O_5) phosphoric acid, which is equivalent to 1.0 ton of P_2O_5. Since 2.7 tons of sulfuric acid (100-percent basis) are required to produce the 1.85 tons of phosphoric acid, the hypothetical producer could pay $11.94/ton (100-percent acid basis) for the sulfuric acid and be able to make phosphoric acid to sell for the same price as that made in Louisiana and shipped to the Cairo area. However, the sulfuric acid would probably have to be offered at somewhat less than $11.94/ton in order to provide incentive to the new phosphoric acid producer. Under

these circumstances, the sulfuric acid could probably be sold for $10 to $11/ton (100-percent acid basis). Since little or no transportation would be required, 78-percent and 93-percent acid should sell for the same price per unit quantity of 100-percent acid.

REFERENCES

D1. McKee & Company, Arthur G., Systems Study for Control of Emissions -- Primary Nonferrous Smelting Industry, Final Report to National Air Pollution Control Administration, June 1969, Contract No. PH 86-68-85.

Table D-1

ESTIMATED SULFURIC ACID DEMAND IN SELECTED PRODUCING AREAS, 1966
(thousands of short tons 100% H_2SO_4)

	Cairo-Paducah	St. Louis	Chicago
Phosphate fertilizers	107	789	1,375
Ammonium sulfate	13	5	113
Iron and steel pickling	--	3	169
Rayon and cellulose films and fibers	--	--	74
Petroleum refineries	10	81	78
Sulfonated detergents	15	57	33
Ammonium sulfate	--	28	12
Titanium dioxide	--	410	--
Hydrofluoric acid	45	--	38
Total Major End Uses	190	1,373	1,892

Source: A. D. Little, Inc.

Table D-2

PHOSPHORIC ACID PRODUCTION COSTS
(per ton 100% P_2O_5)

	Plant Location			
	Louisiana		Cairo, Illinois	
	Quantity Required	Cost/Ton 100% P_2O_5	Quantity Required	Cost/Ton 100% P_2O_5
Phosphate rock (72 bpl)				
$6.00/ton + $3.25 frt	3.2 tons	$29.60		
$6.00/ton + $5.85 frt			3.2 tons	$37.92
Elemental sulfur				
$30/long ton + $1.50 frt	0.82	25.83		
			0	0
H_2SO_4 manufacture $3.40 per ton (excluding cost of sulfur but including 15% return on investment)	2.7	9.18	0	0
Phosphoric acid manufacture (excluding raw material cost and profit)		14.00		14.00
Barge freight $3.00/ton (phosphoric acid from Louisiana to Cairo, Ill.)	1.85	5.55		0
Subtotal		$84.16		$51.92
H_2SO_4 purchase price $11.94/ton (to match cost of producing phosphoric acid at Cairo, Ill. and in Louisiana)		0	2.7	32.24
Total		$84.16		$84.16

Source: SRI